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Suppression of liquid slugs and phase separation through pipeline bends

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Abstract

This study examines the suppression of liquid slugs in the transport and separation of multiphase flows in pipelines. Two well-known slug control approaches are evaluated in this paper. The methods are employed with the aim to control and stabilize an undesired and unstable flow regime, optimize flow production, reduce operating costs, and improve overall safety requirements of oil and gas pipelines. Unlike designs with an additional flowline to separate gas upstream, this study shows that active topside choking can suppress slugs and stabilize the system flowrates and pressures without the requirement of separation upstream of the topside valve. Careful choking is required to minimize production losses that can result from excess back pressure. A riser-based, gas-lift method reduces system instability and increases production. This study also reveals that negligible improvement in stability is achieved when large volumes of gas are injected. The system shifts into an annular flow regime when the injection is further increased. A large separator may be required to accommodate large gas volumes. This research concludes that gas-lift not coordinated with choking is not effective for slug mitigation through pipeline bends. This study presents and discusses new non-dimensional correlations, including slug control inputs in pipelines such as choke openings, based on new experimental data.

KEYWORDS

severe slugging experiments, subsea fluid transport, pipeline-riser system, slug flows

1 INTRODUCTION

In offshore hydrocarbon production, riser systems are often used for transporting multiphase fluids through pipelines with bends, as this can be more efficient, cost-effective, and safer than other transportation methods. However, severe slugging, due to intermittent cyclical surges of liquid and gas phases of a multi-component fluid system, limits the performance of these systems. Slugging is a common flow pattern in multiphase flows of oil and gas upstream production processes. An unstable multiphase regime exists where the flow rate, pressure, and temperature significantly oscillate, causing a potentially negative impact on downstream processing.[1,2]

Naterer[3] described how liquid slugs in multiphase gas-liquid flows are separated into regions consisting of small bubbles dispersed in the fluid system. Bubbles may coalesce to form larger sizes in layers as they are transported through a pipe. In subsea pipelines, the gas and liquid (such as water and oil) may not be evenly distributed throughout the production wells and risers due to the configuration and operating condition. The liquid and gas travel as a plugged section with a large plug of one phase followed by the other phase through the pipeline.[4,5] This type of irregular flow can result in poor phase separation, reduced production capability, and extra fatigue loads on facilities, thereby shortening device lifespan, accelerating component corrosion, and leading to emergency shut-off of production.[5] Slugging in pipeline-riser systems can also cause significant fluctuations in gas and liquid flow rates.[3,4] This can lead to severe vibrations in the riser, platform trips, reduced separation performance, reduced riser integrity, and a decrease in overall safety.

Deepwater fields and production from marginal offshore fields via subsea tie-backs to existing facilities often lead to slugging. Similar problems may occur in fields approaching their end of life, due to a decrease in gas coupled with an increase in water production. In these situations, large fluctuations in gas and liquid production may cause high oscillations and uncontrolled vibrations, ultimately leading to production deferral, riser integrity/safety issues, and potential abandonment of fields near their end of life.[6] Slug elimination and control have been investigated in several studies.[7,9] Various solutions have been proposed to reduce or eliminate slugging. However, some of these methods simultaneously reduce the flow rates and production. Pedersen et al.[9] reported that, although slugging in flow systems has been widely investigated, many challenges remain unaddressed in relation to cost-effective and optimal control of slugging.

Past experimental studies have focused on the development of correlations for the void fractions, pressure drop, physical pipeline parameters (e.g., pipe diameter), and inclination angles to better understand the sensitivities to slugging patterns.[10] However, correlations that include important input parameters for slug mitigation in pipeline-riser system such as choke opening are scarce. Slugging phenomena are especially sensitive to the void fractions and phase velocities, as well as choke openings, pipeline diameter, flow rate, downstream pressure, and pipeline inclination. Some models are also sensitive to surface tension and gas density. Although empirical models exist for evaluating slugging in certain operating conditions, there is a lack of adequate data and suitable modelling approaches that can accurately predict flow conditions over a range of conditions and pipeline geometries. Most of the existing empirical correlations consider gravity-based separation, which can handle small pockets of slugs through a larger contact area. However, current slug mitigation methods and offshore facilities are usually not equipped with large gravity-based separators.

Various types of slugging phenomena, classifications, descriptions, and elimination methods have been reported.[8,11] Some researchers[4,12] described the development cycle of severe slugging. Although the researchers suggested a varying number of steps in developing severe slugging, the fundamental concept is similar. The following steps may be used to describe slug development stages: (1) liquid level rise at the bend, (2) pipeline blockage, (3) slug growth (elongated bubble in riser), (4) liquid production, (5) gas blowdown, and (6) gas production followed by liquid slugs. The different stages of slug formation and growth in the subsea pipeline-riser system are illustrated in Figure 1.

Fluctuations in large-amplitude pressure and flow rates in a pipeline-riser system were reported by Alvarez et al.[12] and Malekzadeh et al.[13].

Two common methods for slug elimination include choking, by reducing the opening of the fluid outlet on the topside of the riser,[14] and gas lifting,[15,16] which involves injection of gas at the bottom of the riser where the liquid loading is anticipated. Gas lifting is an artificial fluid lifting process that is widely used in the petroleum industry, mostly for enhanced oil recovery. Its challenges include high operating costs, gas handling problems, and overall safety, especially in offshore oil recovery, where a smaller footprint and compact design are required. Manual choking of the topside valve is often preferred to suppress slugging in offshore fields. However, choking also lowers production significantly and may impose severe back-pressure in the pipeline system, resulting in repeated shutdowns and production losses.

Two-phase offshore processing facilities typically use compact designs to accommodate the footprint constraint in offshore platforms. Empirical models that include relevant input parameters for compact designs are limited in the literature. More reliable prediction models of slug frequency, gas volume, slug velocity, and production rate are needed for better characterization of flow, and selection and design of appropriate mitigation methods.[5] Also, several aspects of slug mitigation methods are not well understood. In this paper, new predictive correlations, including slug control inputs such as choke openings, will be developed for the evaluation of slug performance under different control schemes. The models will provide useful tools for controller design and system analysis. This paper will present the experimental facilities and test conditions for the measurements, data acquisition, analysis, and comparison of different flow control methods. Results for the various techniques of slugging elimination will be presented and discussed.

2 FORMULATION OF SLUG FLOW IN PIPELINES

Accurate prediction of slug flow characteristics during fluid transport in pipelines with bends is important for effective design and safe operation. In this section, correlations will be developed to predict the behaviour of slugs in pipelines with bends. The functional forms of the correlations will be similar to past studies conducted by others but will expand upon key variables for the equipment design.

Nicklin et al.[17] presented an empirical relationship for slug bubbles in horizontal pipes. The expression for the flow velocity (the velocity of the slug bubble front) is given by the following:

$$u_{sb} = C_0 u_{ls} + C_1 (gD)^{\frac{1}{2}} \tag{1}$$

where C_0 is a weighted velocity distribution parameter, u_{sb} is the weighted mean velocity of the gas phase relative to the liquid phase, u_{ls} is the velocity of the liquid slug, D is the pipeline diameter, and C_1 is a constant of proportionality. The subscript sb denotes slug bubble. Naterer[3] presented the slug rise velocity in a simpler form:

$$u_s = 0.345\sqrt{gD} \tag{2}$$

where u_s is the relative velocity of the slug front relative to the fluid mixture in the pipeline. This relationship can be used for horizontal pipes with low viscosity fluids.

Abdul Majeed et al.[19] presented the following correlation for the liquid slug hold-up, α_{Ls}, in terms of the inclination angle, β, ratio of the gas to liquid viscosities, x, and mixture velocity, U_m:

$$\alpha_{Ls} = (1.009 - (0.006 + 1.3377x)U_m)(1 - \sin\beta) \tag{3}$$

The slug period (T) can be related to the mixture velocity (U_m) as shown below[4]:

$$T = e^x {U_m}^{-y} \tag{4}$$

where

$$x = 0.415\ln(U_{sl}) + 5.339 \tag{5}$$

$$y = 0.072\ln(U_{sl}) + 1.390 \tag{6}$$

In Equations (5) and (6), U_{sl} is the liquid slug velocity.

Schmidt[14] also presented an empirical correlation of the slug period as a function of the flow rate:

$$T = a{U_m}^b \tag{7}$$

where a and b are independent of the gas and the liquid flow rates. The author examined data for various slugging types and showed that the slug period varies linearly with the mixture flowrate. He found that the correlation's accuracy is ±30%, which is in relatively good agreement with the experimental data. The exponent of U_m was $b = -1$; therefore, we have the following:

$$f_s = a{U_m}^{-1} \tag{8}$$

Hill and Wood [16] also proposed a correlation for the slug frequency (f_s) as follows:

$$f_s D = 0.27510^{2.68H_L}U_m \tag{9}$$

The mixture velocity is given by w_{mix}/A, where w_{mix} is the mixture flow rate of the gas and liquid phases, and A is the cross-sectional area of the pipeline.

Although past studies have developed correlations to address different aspects of slug mitigation measures, there is still a lack of predictive models that include the choke opening (a key variable for offshore applications where choking is widely used). In this paper, correlations of slug flow characteristics will be developed based on new experimental data and the previous functional forms of correlations. The correlations will be presented and analyzed in non-dimensional form.

3 EXPERIMENTAL FACILITY

Experiments were conducted to compare various slugging suppression techniques and develop criteria for applying any chosen suppression method. The experimental runs were conducted to emulate slugging in pipe bends and evaluate how the actuators (choke or gas-lift) considered may be applied to suppress the slugs and phase separation at the bends. These will examine how the actuators can impact the transport and separation of multiphase flows in pipelines with bends.

The experiments were conducted in a multiphase flow facility of the Offshore Energy Laboratory at Aalborg University, Denmark (see Figure 2). The internal pipeline diameter, total length, and inclined section are 0.054, 42, and 12 m, respectively, while the riser total length is 6.1 m. The experimental apparatus has pipes made from transparent PVC to enable flow patterns in the piping to be visually observed. A liquid phase (water and mineral oil) is supplied into the system from buffer tanks by centrifugal pumps. A centrifugal water feed pump with an operating condition of 1 L/s at 162.7 m and a maximum pressure of 25 bar (2.5 MPa) is used to supply water into the system. The water and oil flow into a mixer with a Venturi-based design. Each supply pipe is equipped with check valves for safety to prevent back-flows. The outflow is received into a downstream separator.

The pipeline-riser-separator system consists of a complete flow-loop. The flow measurements are collected by flow transmitters, Coriolis flow metres. The setup collects the pressure measurements using pressure transmitters (PTs) installed along the test pipeline-riser system loop. The PTs have operational ranges of 0 – 16 bar (0 - 1.6 MPa)and a piezo-resistive measuring cell with a ceramic diaphragm. The pressure transmitter is noted for its fast step response time of fewer than 5 ms.[20]

The test conditions were designed based on similar tests conducted by other researchers in past studies.[20,21] Slugs were created with constant liquid and gas flow rates of 0.4 and 0.000484 kg/s, respectively. These values correspond to past data reported in previous studies.[1,20] A constant air injection rate was maintained into the system through the inlet of the pipeline at a constant pump pressure of approximately 1.8 bar (0.18 MPa). A constant inlet pressure was achieved by maintaining a constant pump speed with a proportional-integral controller. The gas controller was used to maintain a constant gas flowrate through a constant gas volume fraction (GVF). For the gas-lift case, an additional 0.625 nm^3/h (normal metre cube per hour) of air was injected into the flow system every 5 min through the riser base until a maximum capacity of 5 m^3 was reached.

The reservoir section accommodates the inflow supply equipment, including the mixing of water, oil, and gas at the pipeline inlet. The flow equipment is shown in Figure 2, where the outlet fluid moves into the 3-phase separator. Steady-state operating conditions were obtained for each valve opening. The valve openings are in the range of 10%-100%, which covers the majority of actual operating conditions for fluid recovery in typical oil and gas processing applications. During each experiment, the choke degree opening is stepped down by 10% after a set period. First, a slug was created during the first 50 s. The slug dominated flow is then allowed for 700 s for the 100%-degree choke opening. During the subsequent openings (90%, 80%, and so forth, down to 10%), the duration was 300 s for each of the degree openings.

The slug flow was characterized by pressure and flowrate oscillations in the pipeline. Under stable flow conditions (continuous gas penetration where the gas penetration is non-zero), there is no liquid blockage at the bottom of the riser. The flow is steady with bubble or small slug pockets. The flow rate used to create the stable slugs was maintained within the slug regime throughout each experiment. For each slug elimination scheme considered, a slugging flow regime was created in the pipeline until a fully developed slug occurred. Similar to past studies,[1,20] all data acquisition and controls were performed with Simulink Real-time models for real-time simulations. Multiphase flow in the system consisted of gas and liquid phases. The liquid was transported through the pipeline-riser to the separator, while compressed air was injected into the pipeline at the bottom of the riser.

In total, 19 experiments (10 choking method and nine gas injection method) were conducted. For each experiment, gas and liquid inflow conditions were controlled by means of pumps and the GVF. For each test, measurements were taken with the aforementioned instrumentation at different locations in the flow loop. The sampling frequency was 100 data points per second for all variables, each lasting about 1.5 h with approximately 520 000 measurements per test. Each new experiment was repeated to ensure reproducibility of the tests. The separator pressures were kept at atmospheric conditions. The temperature, measured with temperature transmitters (TT) installed along the test loop, was relatively constant at 20^{o}

4 EXPERIMENTAL UNCERTAINTIES

Uncertainties associated with the reported measurements were investigated by the method of Kline and McClintock.[22] The measurement errors in each experiment were classified as bias (B) or precision (S) errors. Bias errors were estimated from calibration procedures based on curve fitting of calibrated data. The measured and manufacturer prescribed uncertainties are presented in Table 1. The uncertainties include variables determined from direct sensor measurements, which are largely dependent on the equipment accuracy. Uncertainties for the measurement equipment were obtained from the manufacturer's data sheets of fixed error estimates for the flow rates, density, temperatures, and pressure sensors. The calculated uncertainties associated with other quantities, such as GVF and mixture velocity, were obtained by substituting average values of the measurements into the appropriate equations (also shown in Table 1).

Procedures for reporting measurement uncertainties for single measurements, such as pressures and flow rates, were previously presented by Adeyinka and Naterer,[23] Kline and McClintock,[22] Moffat,[24] and others. Errors propagated by an individual measurement can be evaluated separately using sensitivity coefficients, which involve the measured variables. Overall uncertainties for single measurements can be determined by the following propagation equation[22,24]:

$$U_{0.95} = \left[\left(B_{x_i}\right)^2 + \left(S_{x_i}\right)^2\right]^{\frac{1}{2}} \tag{10}$$

Since calibration is considered in the measurements, the fixed errors represent the bias, while the random errors due to measurement variations are determined by the precision limit calculations.

The precision errors are estimated from the unsteadiness in the measuring process. The unsteadiness is determined through repeated experiments. These measurement errors are affected by the measurement system and spatial variations in the measured variables. The precision limit of an individual measurement, x_i, such as pressures at different locations in the pipeline-riser, flowrates, and mixture density are calculated from the measured data based on the following equation:

$$S_{x_i} = \frac{t\sigma}{N} \tag{11}$$

where S_{x_i} is the precision error, t is the confidence coefficient for an experiment within a 95% confidence level, N is the number of measurement samples, and σ represents the standard deviation determined from the measured data as follows:

$$\sigma = \left\{\frac{1}{N-1}\sum_{i=1}^{N}(X_i - \bar{X})^2\right\}^{\frac{1}{2}} \tag{12}$$

The precision limit of the mean measurement, $\bar{X}$, of the sample, N, is determined by the following:

$$\bar{S}_{x_i} = \frac{1}{N}\left\{\sum_{i=1}^{N}\frac{X_i - \bar{X}}{N-1}\right\}^{\frac{1}{2}} \tag{13}$$

The total uncertainty ($B + S$), for single measurements such as T, P_s, P_p, P_{in}, w_l, w_g, and w_{mix} are determined from the method of Kline and McClintock[22]:

$$\psi_{0.95} = \left\{\left(B_{x_i}\right)^2 + \left(2\sigma S_{x_i}\right)^2\right\}^{\frac{1}{2}} \tag{14}$$

Typical values of the standard deviation are estimated for the production rate, pipeline pressure, pressure at the bottom of the riser, and pressure at the top of the riser as 14.3%, 0.04%, 9%, and 0.4%, respectively. The values have the precision limits of 0.4%, 0.002%, and 0.02%, respectively.

The total uncertainty ($B + S$) of the pressure measurements along the pipeline-riser setup is calculated as follows:

$$\psi_{P_i} = \sqrt{S_{P_i}^2 \pm B_{P_i}^2} \tag{15}$$

The total uncertainty of the flow rates and temperature are calculated in the same manner by replacing the pressure components with flow rates and temperatures, respectively. The pressures, temperatures, and the flowrates then become the following:

$$P_i = \overline{P_\iota} \pm \psi_{P_i} \tag{16}$$

$$T_i = \overline{T_\iota} \pm \psi_{T_i} \tag{17}$$

$$w_i = \overline{w_\iota} \pm \psi_{w_i} \tag{18}$$

where i is used to denote the position in which the measurements were taken, such as the top of the riser, and the bottom. The mean variables, $\overline{P_\iota}$, $\overline{T_\iota}$, and $\overline{w_\iota}$ are determined as follows:

$$\bar{X}_i = \frac{1}{N}\sum_{i=1}^{N} X_k \tag{19}$$

This procedure is used to determine the uncertainties associated with calculated variables such as the GVF and the mixture velocity. A more detailed description of the measurement uncertainties is given by Igbokwe.[25]

In Table 1, the calculated uncertainties associated with the liquid and gas flow rates, and top and bottom pressures, are reported. When the appropriate values are substituted in the uncertainty equations, the uncertainties of the liquid pipeline pressure, and pressures at the top and bottom, remained relatively constant at ±1.78% of the liquid flow rate value; $\pm 5.56 \times 10^{-4}$ or ±8.34% of the gas flow rate; ±0.011 or ±0.63% of the pipeline pressure; ±0.01 or ±0.85% of the pressure at the top of the riser; and ±0.01 or ±1.08% of the pressure at the bottom of the riser. The worst-case uncertainties are associated with the gas volume fraction measurements and may be attributed to slugging behaviour. The gas volume fraction continued to change throughout each test, as a result of the various stages in the slugging process, including cyclical gas accumulation.

5 CASE 1 – ACTIVE CHOKING AT THE TOP OF THE RISER

Consider a case of open-loop stepwise choking of the topside valve. In order to examine the impact of choking on slugging and fluid recovery, the choke valve opening was stepped down by 10% for each test, from fully open (100%) down to 10% open. The lower limit is below the bifurcation point, or the point of choke opening in which the system shifts from slugging to non-slugging flow behaviour. The separator pressure can increase the pipeline operating pressure and act as an anti-slug control mechanism. Increasing the pipeline pressure decreases the volume of gas in the pipeline and increases the pipeline liquid holdup.[26]. The increased backpressure also lowers the gas velocity and enhances gas-liquid mixing. With the higher liquid cut in the pipeline and reduced gas velocities, a continuous mixture flow is maintained in the system.

In this case study, the separator pressure is maintained at atmospheric pressure. This ensures that results are solely impacted by topside choking. Constant liquid and gas flow rates of 0.4 and 0.00048 kg/s, respectively, are maintained in the system through a 0.054 m inlet pipeline. The fluid densities and viscosities are approximately 900 kg/m^3 and 9.04×10^{-2} kg/m/s^2 for the liquid phase and 1.988×10^{-5} kg/m^3 and 0.000181 Pa·s for the gas phase, respectively. These values are typical fluid properties obtained at standard conditions as the temperature dependent fluid properties did not vary significantly throughout the experiment. Temperature was maintained close to atmospheric condition at an average value of 20.4°C. The fluid properties above represent averaged values of the gas phase (air) and liquid obtained at atmospheric conditions. Constant inflow rates were maintained at a constant average pump pressure of 1.8 bar (0.18 MPa).

Experimental results are presented in Figures 3-7. The results show input / output relationships and how they affect the system operating conditions. Liquid and gas inflows are related to the mixture flow at the outlet of the separator, which is controlled by the topside choke valve. Figure 3 shows that the separator pressure was maintained at atmospheric conditions, while the riser base injection nozzle was off, after which the choke valve was stepped down from fully open (100%) to 10%.

Minor fluctuations in pressure were measured as a result of the back pressures created when valve openings were reduced. If the inflow conditions, liquid and gas flow rates, outflow condition, and separator pressure are kept constant, then the system flow responses for various choke openings can be evaluated. The gas injection was zero throughout this case study.

In Figure 4, a GVF fluctuation occurs due to slugging. The fluctuation amplitude is higher in the slugging region. The fluctuations challenge the controller because the actuator's time constant is much shorter than the pump or the influence caused by slugging. In other words, it reacts rapidly to any disturbance, while this is not the case for the pump since the pump dynamics are much slower than the compressor dynamics. However, the gas controller manages to keep the subsequent oscillations to a minimum so the GVF is close to constant (Figures 3 and 4).

The pump pressure in Figure 4 is more uniform because the induced back-pressure added from the topside choking is minor compared to the pump controlling the pressure. It is also stepped over long periods, which gives the pump time to stabilize.

Figure 5 reports the measured average gas inflow and liquid inflow rates. The inflow rates show the flow responses based on the system flow condition.

Due to the oscillating system pressures and mass flow rates, the inflow performance varies, particularly for larger choke openings, where the fluctuations are highest. At lower choke openings, the inflow variables are nearly constant since the system tends to a stable flow regime. Flow stabilization in the non-slug region impacts both the liquid and gas flow rates. The pressure and flowrates stabilize after about 4000 s.

In Figures 5 and 6, the slug is eliminated at about 4000 s. Table 2 lists the production at each step size of choking. The table compares the impact of slugging on fluid recovery at the process facility outlet of the choke valve. At larger openings, the production rates are higher than smaller choke openings. However, the slugging frequency is more intense compared to lower choke openings.

Figure 6 reveals that fluid recovery at the choke valve decreases close to the stabilization point when the choke opening decreases. This production loss is also shown in Table 2. Acceptable production rates may not be reached for particular choke openings when the system stabilizes in a real application. Generally, acceptable production rates depend on the operator's objectives, lifespan of the well, and economic considerations. Choking may not be most beneficial for all situations where high production rates are most desired and when other mitigation approaches are viable. Other methods, such as robust feedback control, can be used to increase the flow rate.

Table 2 compares the impact of slugging on the flow conditions. A sharp change is noticed between the slugging and non-slugging choke openings. Each slug cycle has an average of 87 s for choke amounts between 100%-70%. The period doubles between 40%-30% openings. In non-slugging flow (<20%), the period is infinite, and the back-pressures also increase at a smaller choke opening when the system stabilizes. Figure 7 shows that the density fluctuates over time during the slugging. A lower limit is observed during the flow initialization stage at about 0.5 kg/m^3. The slug frequency, however, decreases as the system stabilizes. The decreases in flow, pressure, and density variations under non-slug gas flow are an indication that the gas bursts from each slug until it is eliminated.

In the choking technique, the choke valve is manipulated to add back-pressure and modify the GVF. When the choke is sufficiently decreased, the slug flow turns into bubble flow, in which bubbles are immersed with the continuous liquid phase in the riser. The choking technique controls the slug flow by suppressing the development, or the growth, of liquid slugs at the base of the riser. During the initial slug formation, when the liquid blocking starts, the volume flow rate and pressure in the riser decrease. This differential pressure actuates the valve to maintain a set volumetric flowrate. Consequently, the pressure downstream of the slug formation decreases and liquid slugs are pushed upward by the pipeline pressure.

6 CASE 2 – GAS INJECTION SCHEME

The second case study considers gas injection with a fully open choke valve. Similar to the previous case, constant liquid and gas flow rates of 0.4 and 0.00048 kg/s, respectively, are maintained through the 0.025 m inlet pipeline. Gas injection is introduced at the bottom of the riser. The fluid densities and viscosities are 900 kg/m^3 and 0.090445 kg/m/s^2, respectively, for the liquid phase, and 1.988×10^{-5} kg/m^3 and 0.000181 Pa·s, respectively, for the gas phase. The constant inflow rates were maintained with an average pump pressure of 1.8 bar (0.18 MPa). Minor fluctuations in pressure are observed as a result of the back-pressures introduced when the valve openings are reduced. Gas is injected at the bottom of the riser where the liquid is accumulated to ensure continuous gas penetration into the riser.

Gas is injected at 0.625 nm^3/h increments every 300 s until the maximum capacity of 5 nm^3/h is reached. Additional gas injection into the riser base increases the total volume of gas into the system, thereby resulting in higher flow velocities. The results from the gas-lift experiments are shown in Figures 8 – 10. The results are collected over the period of 3000 s for the experiments. Gas lift affects the pressure at the bottom of the riser, top pressure, production rates, and frequency of the slugs. The experimental results show that an improvement in the system stability is relatively small before large volumes of gas are injected. Before stability can be achieved for the gas lifting technique, the system flow changes into an annular flow regime. The gas lift inability to obtain system stability is observed in Figures 9 and 10.

Figures 8 and 9 show that the maximum gas injection capacity is reached without a substantial change in the pressures and flow rate responses. According to Figure 10, slugging is not eliminated for the gas lift scenario under the given test conditions.
This affects operating costs and increases gas handling problems, as reported previously by Taitel.[4] A larger separator is required to accommodate large gas volumes and unstable slug production.

This slugging behaviour is also demonstrated in Table 3. Both the flow rates and pressures continue to produce high amplitudes throughout the test period. The fluctuations in the system are depicted with both high frequencies and high amplitudes. Although small reductions in the amplitude are briefly observed around 2500 s, the oscillations are not significantly reduced when the maximum injection capacity is reached. Similar to previous studies,[15,16, 27] these results show that large volumes of gas are required to reduce the severity of slugging.

A large separator or slug catcher is required to accommodate both the large gas volumes and unstable slug productions. The riser-based gas-lift method can reduce system instability and increase production (see Table 3 and Figure 12). However, this is not suitable for offshore applications due to cost, space, and weight aspects. Gas lift mainly increases the fluid velocity in the riser. The gas flow eventually dominates the riser flow as gas injection continues, before stability can be achieved. The significant positive impact on the system includes a reduction of system pressure and slugging frequency. The gas lift also reduced the severity of blowout of liquid in the pipeline and allowed a continuous steady flow in the flowline and riser. This occurs because the injection gas aerates the hydrostatic column, thereby reducing the force required to push the fluid to the surface, and hence allowing for smoother fluid recovery downstream. In a blowout (sudden and uncontrolled flow) situation, there is uncontrolled fluid production downstream, which strains the fluid processing equipment and impairs the fluid handling capabilities of downstream devices.

Schmidt et al.[4,14] used riser base injection and reported gas handling challenges of compression and pressurization. Also, a Joule-Thompson effect may aggravate the flow conditions by making the transported gas susceptible to wax precipitation and hydrate formation.[28] Measured results show that the gas injection rate to eliminate slugging is higher than the flow rate of gas in the pipeline (Figure 11). Gas injection reduces the pipeline pressure and can boost production. It is also beneficial for well and riser unloading to enhance start-up conditions and improve production stability.

In gas lift applications, the gas injected at the bottom of the riser reduces the hydrostatic head imposed by a liquid slug in the riser column. Hence, the pressure of the pipeline decreases.[12] Past studies have shown that gas lifting increases the velocity and reduces liquid holdup in the riser.[16,29,30] The injected gas also helps to carry liquid to the surface receiving facilities. When the gas volume in the system is sufficient to ensure continuous fluid lifting, a stabilized flow can be achieved. Large volumes of gas are required to achieve stability and eliminate slugging. This also increases the potential for gas handling challenges at the surface processing facility, as well as energy costs.[16] Insufficient injection gas can increase slugging frequency.[30,31] Compressor requirements and costs are also challenges in gas lifting as a slug elimination technique. The current results indicate that no substantial improvement in stability is obtained when large volumes of gas are injected, although significant energy is consumed.

7 CORRELATION RESULTS AND DISCUSSION

In this section, new correlations are presented for the slug flow rate, velocity, slug frequency, topside and riser bottom pressures, and gas injection rates. These will be represented in terms of the choke opening fraction, z, and gas injection rates. Measurements have been taken for pressures at the inlet and throughout the pipeline, and at the bottom and top of the riser. The separator pressure, fluid densities, temperatures, and flow rates at the inlet and outlet of the setup were measured for different gas injections and choke openings. The experimental results will be used in this section to develop new correlations of key system design variables.

A plot of the dimensionless production rate, $w_m/w_{m,max}$, against the percentage choke opening, for a range of choke openings (100%-10%, in 10% decrements) at a given diameter is shown in Figure 13. The data cluster for high choke sizes (choke openings above 42%) indicates repeated oscillations at high amplitudes. For the choke openings below 42%, the data are more uniformly distributed.

From Figure 13, an approximately linear relationship exists between the production rate and choke opening. A coefficient of variation (r^2) and gradient of 0.88 and 0.72 are obtained, respectively. This form of the correlation is similar to previous forms presented by Nicklin [17]. The system variation in the amplitudes for the stable region below the bifurcation choke opening may be attributed to the valve dynamics.

The correlation is expressed as a function of the pipe diameter and choke size opening, which have a linear relationship with the slug production at the outlet. The production rate varies linearly with the choke opening. The flow through the choke is maximum for a 100% opening, and lowest for the 10% opening. The r^2 value of 0.88 indicates that the flow out of the choke is well correlated to the percentage choke opening. From the experimental data, the mixture produced from the choke outlet can be correlated as a function of the percentage choke opening:

$$w_{mix} = 0.723z \tag{20}$$

where w_{mix} is the mixture flow rate at the topside of the choke valve, while z is the choke opening percentage. Equation (2) may be expressed in terms of the slug flow velocity at the topside of the choke based on the cross-sectional area open to the flow, $u_{Sl} = 0.567D^2z$, where u_{Sl} is the slug mixture flow velocity at the topside of the choke valve, D is the pipe internal diameter, and z is the choke opening percentage.

Frequency data for the analysis was obtained from measurements over 500 s at a sampling rate of 100 Hz. The sampling rate is oversampling as the riser-induced slugs have lower frequencies. However, it is used to guarantee that all dynamics are captured such as trends during the slug. Figure 14 shows the normalized frequency plotted against the choke opening. It is shown that a logarithmic relationship exists between the slug frequency and choke opening.

A logarithmic regression analysis was used to obtain r^2, $\underline{a}$, and b, as 0.87, 0.35, and 0.864, respectively (Figure 14). The slugging frequency can provide key information in the design of the appropriate controller bandwidth based on the flow conditions.

The final form of the correlation for the estimation of the slug frequency is given in a logarithmic form based on previous studies. The slug frequency is a modified form of the correlation of Schmidt et al.[14] The final correlation for estimating the frequency of slugging is expressed as a function of the choke opening as follows:

$$f_{slug} = 0.864 + 0.351\ln(z) \tag{21}$$

Figure 12 shows that the frequency of slugs decreases as the choke opening decreases, although the slug frequency is relatively constant for choke openings between 100%-60%. The frequency decreases for smaller choke openings until slugging is eliminated for choke openings of about 24%. This phenomenon indicates that when more liquid slugs are produced at the choke outlet, the gas and liquid velocities become higher.

A plot of the dimensionless pressure at the bottom, $p_b/p_{b,max}$, and top, $p_t/p_{t,max}$, of the riser against the choke opening is presented in Figure 15. The correlations were based on 5000 data points obtained from the experiments. The general form of the correlation for the pressure is ax^b for the bottom pressure and $a\ ln\ (x) + b$ for the top pressure. After a regression analysis, the values of r^2, a, and b are 0.50, 0.87, and -0.01, respectively, for the bottom pressure, and 0.39, 0.01, and 0.86, respectively, for the pressure at the top of the riser.

The gas injection rate is an input parameter in the gas lift design and compression power requirements. A plot of the gas injection ratio (ratio of inlet to total gas flow rates) is illustrated in Figure 11. This shows the volume of gas injection requirements as a function of time. The results indicate that gas injection should be maintained to continue liquid production and prevent slug development. Although the gas lifting method has increased the fluid production (Figure 16), stability was not reached after large volumes of gas were injected (Figures 9 and 10). It is also shown that a significant amount of energy is consumed (Figure 8).

From the analysis of the gas-lift data, a general form of the gas requirement parameter is $a\ ln\ (x) + b$. The coefficients of variation (r^2), a, and b, are obtained for the dimensionless gas injection as 0.91, 0.23, and -0.14, respectively. The regression shows that the functional variables are linearly correlated to the injection rates. The final correlation to estimate the volume of gas required to maintain a slug free system is given by the following relationship:

$$w_{Ginj} = (0.0239 \ln(t) - 0.01382)\alpha w_L \tag{22}$$

where w_{Ginj} is the gas injection rate required to keep the system in a non-slugging region, α is the void fraction, and w_L is the liquid flow rate into the pipeline, which is equivalent to the liquid production rate at the wellhead of a production platform.

A correlation for the production rate with gas lifting is primarily used for anti-slug control. From a linear regression analysis, 0.07, 0.58, and 0.05, respectively, are obtained for a, b, and r^2. The general form of the correlation is linear ($ax + b$). Figure 12 shows the dimensionless fluid production, $w_m/w_{m,max}$, at the choke outlet, correlated as a function of time for the gas injection method. The results indicate that the total fluid recovery increases as the injected gas volume rises. The production rate increases slightly and becomes constant. The general form of the correlations and their coefficients are summarized in Table 4.

8 CONCLUSIONS

In this paper, active choking and gas lift techniques were examined to control slugging in pipelines with bends. Dimensionless correlations were developed for estimating key design parameters for slugging mitigation. Results were presented for active choking of the topside valve, and gas lifting at the base of the riser. Choking eliminates slugs by inducing a back-pressure in the upstream pipeline. The method is found to be cost efficient and it stabilizes system flow behaviour by suppressing riser-induced slugs and effectively controlling slugs that propagate through the horizontal pipelines. The results show that choking stabilizes the fluctuations in pressures and flow rates. In addition, riser base gas injection reduced the system instability and increased production but had disadvantages of larger space and weight requirements, as well as significant capital and maintenance costs. Based on the results, large separators or slug catchers may be required to accommodate the gas volumes and unstable slug productions propagating at high liquid and gas velocities. New non-dimensional correlations that include slug control inputs in pipelines, such as choke openings, have also been presented.

AUTHOR CONTRIBUTIONS

This study was conceptualized and coordinated by G. F. Naterer as the principal investigator. Both G. F. Naterer and S. Zendehboudi provided technical guidance and directed all aspects of the research. L. C. Igbokwe, S. Pedersen and S. Jespersen designed the experiments and collected the data. L. C. Igbokwe, G. F. Naterer, S. Zendehboudi and S. Pedersen performed data analysis. L. C. Igbokwe drafted the manuscript. All authors contributed to manuscript revisions.

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NOMENCLATURE

Variables

A	Pipe cross-sectional area (m^2)
B	Bias limit
D_p	Pipeline diameter (m)
D_r	Riser diameter (m)
f_s	Slug frequency (Hz)
g	Gravitational acceleration (m^2/s)
GVF	Gas volume fraction
h_L	Liquid film height (m)
L_r	Riser length (m)
L_p	Pipeline length (m)
P	Pressure (Pa)
R	Specific universal gas constant (J/ (kg K))
S	Precision limit for single measurement
T	Temperature (K)
U_m	Mixture velocity (m/s)
w_{mix}	Mixture flowrate (m^3/s)
U_{sg}	Superficial gas velocity (m/s)

U_{sl} Superficial liquid velocity (m/s)

Z Riser-top choke opening

Greek symbols

ψ Uncertainty

α Void fraction in the pipeline

θ Inclination angle of the pipeline

ρ_g Gas density (kg/m^3)

ρ_l Liquid density (kg/m^3)

Subscripts

r Riser

p Pipeline

m Mixture

s Slug

sb Slug bubble

i Inlet

t Top

b Bottom

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Figure Captions

GRAPHICAL ABSTRACT In offshore hydrocarbon production, riser systems are often used for transporting multiphase fluids through pipeline bends. These bends and subsea tie-backs to existing facilities increases the potential for severe slugging, thereby limiting the performance of these systems. Active topside choking is found to best control and stabilize slug flows without the requirement of separation upstream of the top choke valve. New non-dimensional correlations, including slug control inputs in pipeline such as choke opening size, are developed based on new experimental data

FIGURE 1 Stages of development of slugging: (1) liquid level rise at the bend; (2) pipeline blockage; (3) slug growth (elongated bubble); (4) liquid production; (5) gas blowdown; and (6) gas production followed by liquid slugs

FIGURE 2 Experimental facility for slugging mitigation

FIGURE 3 Case 1 - Measured system inputs showing no gas support and constant separator pressure (atmospheric) while the choke valve is stepped down from 100% to 10% opening

FIGURE 4 Case 1 - Measured flow controller responses as a function of time

FIGURE 5 Case 1 - Measured average inflow at the pipeline inlet against time

FIGURE 6 Case 1 - Measured output variables for the choke test over the experiments

FIGURE 7 Case 1 - Fluid densities for stepped choke valve testing

FIGURE 8 Case 2 - Measured system inputs when the choke is fully open and the air-water mixture is separated by a gravity separator under atmospheric conditions

FIGURE 9 Case 2 - Measured average system flow conditions for the gas injection case

FIGURE 10 Case 2 - Measured average pressure at the top of the riser and the measured average riser-base pressure with time

FIGURE 11 Normalized gas injection requirement for slug elimination vs. time with gas lift (a= 0.234, b=0.138, r^2=0.91)

FIGURE 12 Production rate vs. gas injection ratio (a= 0.07, b=0.59, r^2=0.12)

FIGURE 13 Measured dimensionless average production rate at the choke outlet for various choke percentage openings over an average test period of 4300 s (a= 0.72, b=0, r^2=0.88)

FIGURE 14 Normalized average frequency of slug as a function of the percentage choke opening at choke sizes between 100% and 10% (a= 0.352, b=0.864, r^2=0.87)

FIGURE 15 Average pressures measured at the bottom of the riser (a= 0.87, b=0, r^2=0.5) and top of the riser against percentage choke opening (a= 0.01, b=0.4, r^2=0.5)

Table Captions

TABLE 1 Bias, precision, and total uncertainties of measured results

Variable	Averaged value, $\overline{X}$	Bias (±)	Precision (±)	Uncertainty, $\psi_{0.95}$ (±)	Uncertainty (% value)
w_l (kg/s)	0.61	5.56×10^{-4}	0.011	0.01	1.78
w_g (kg/s)	6.82×10^{-4}	5.56×10^{-5}	1.20×10^{-5}	5.69×10^{-5}	8.34
P_{in}, bar (KPa)	1.76 (176)	0.01 (1)	2.37×10^{-3} (0.24)	0.01 (1)	0.63
P_p bar (KPa)	1.17 (117)	0.01 (1)	4.79×10^{-4} (0.048)	0.01 (1)	0.85
P_b bar (KPa)	1.71 (171)	0.01 (1)	3.01×10^{-3} (0.30)	0.01 (1)	0.61
P_{top}bar (KPa)	0.93 (93)	0.01 (1)	1.6×10^{-4} (0.016)	0.01 (1)	1.08
P_s bar (KPa)	1.03 (103)	0.01 (1)	3.7×10^{-4} (0.037)	0.01 (1)	0.97
w_{mix} (kg/s)	0.53	5.56×10^{-4}	0.005	0.01	0.91
$T(K)$	293.36	0.01	4.51×10^{-5}	5.41×10^{-5}	1.84×10^{-5}
ρ_{mt} (kg/m^3)	0.87	0.001	0.004	0.004	0.46
GVF	1.55×10^{-3}		9.35×10^{-4}	9.35×10^{-4}	39.78
ρ_g (kg/m^3)	2.09	7.27×10^{-5}	1.54×10^{-7}	7.28×10^{-5}	0.004
U_m (m/s)	0.53	4.75×10^{-4}	9.26×10^{-4}	4.83×10^{-4}	0.09

TABLE 2 Impact of slugging on various choke openings

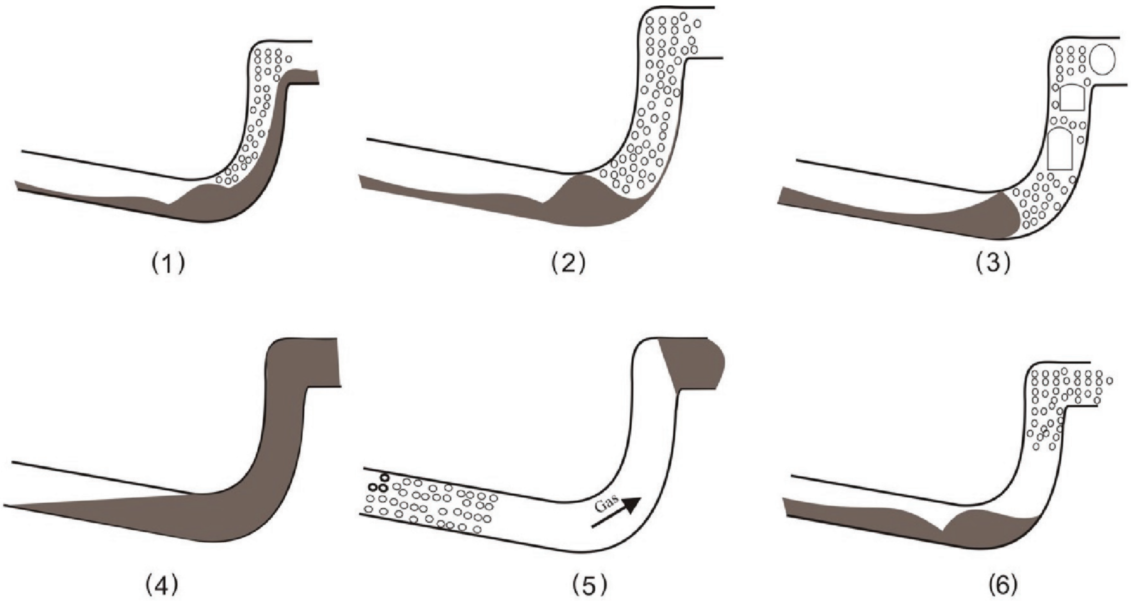
Choke opening (%)	Production rate (kg/s)	Slug period (s)	Slug
100	0.50	87.13	Yes
90	0.54	85.46	Yes
80	0.52	87.8	Yes
70	0.51	89	Yes
60	0.46	97	Yes
50	0.38	108	Yes
40	0.30	134	Yes
30	0.19	217	Yes
20	0.10	infinite	No
10	0.06	infinite	No

TABLE 3 Production flow rates for the gas lift scenario

Gas injection (kg/s)	Production flow rate (kg/s)	Slug period (s)	Slugging
1.25×10^{-6}	0.449	78.5	Yes
0.125	0.635	83.6	Yes
0.249	0.629	84.1	Yes
0.375	0.639	86	Yes
0.499	0.662	83	Yes
0.625	0.638	82	Yes
0.750	0.637	84	Yes
0.875	0.629	83	Yes
0.999	0.628	86	Yes

TABLE 4 Coefficients of correlations of the experimental studies

Relationship	**Correlation**	**R^2**	**a**	**b**
Production rate vs. choke opening	ax + b	0.88	0.71	0
Slug frequency vs. choke opening	a ln (x) + b	0.87	0.35	0.86
Gas injection required over time	a ln (x) + b	0.91	0.23	-0.14



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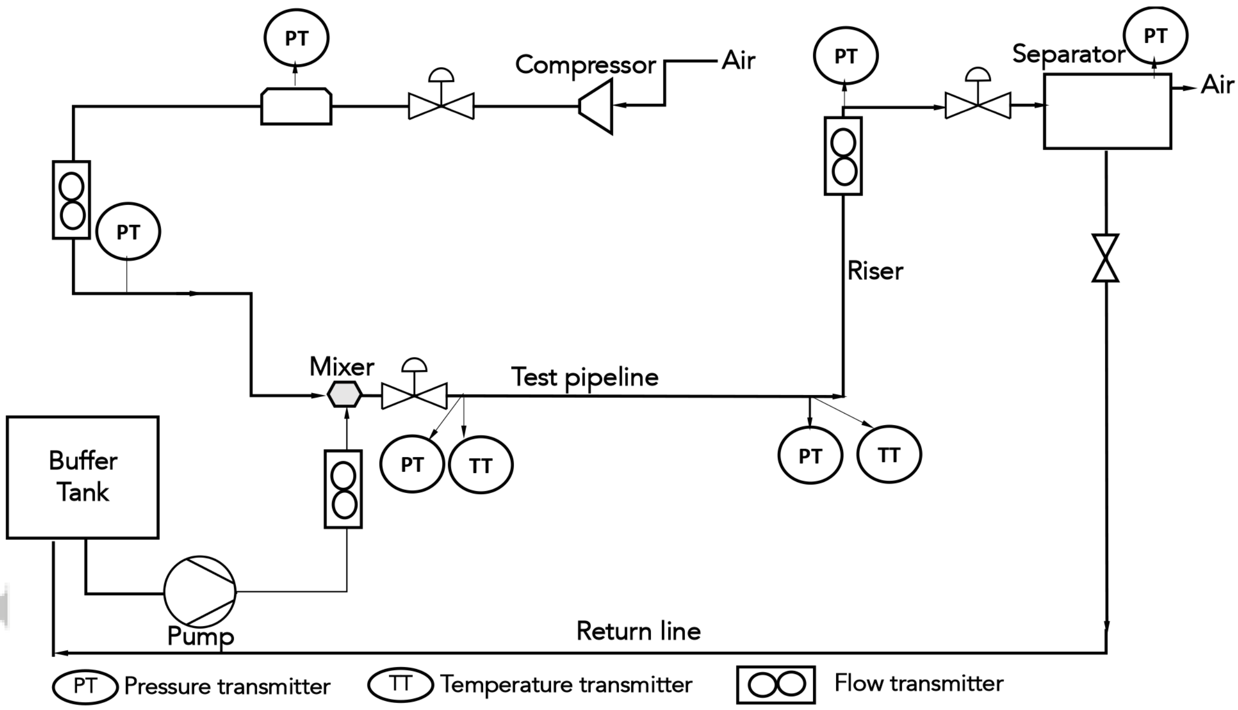

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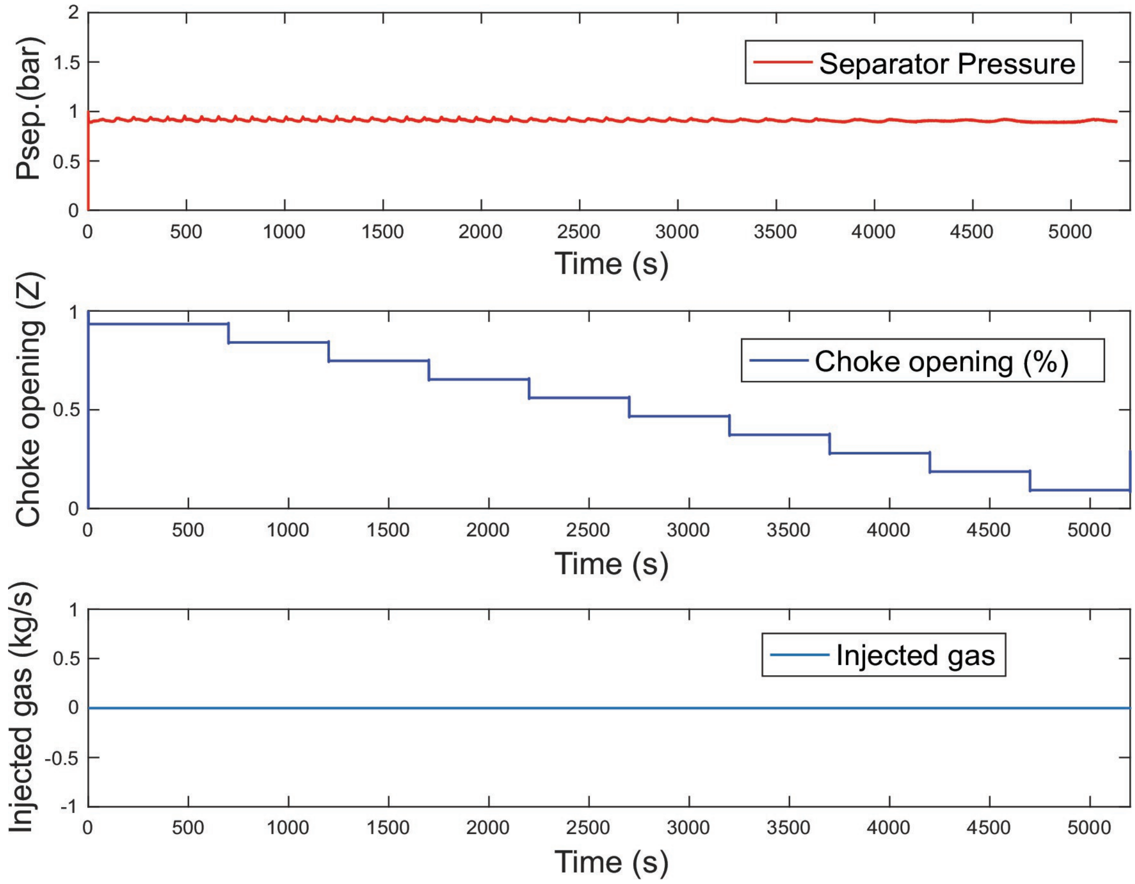

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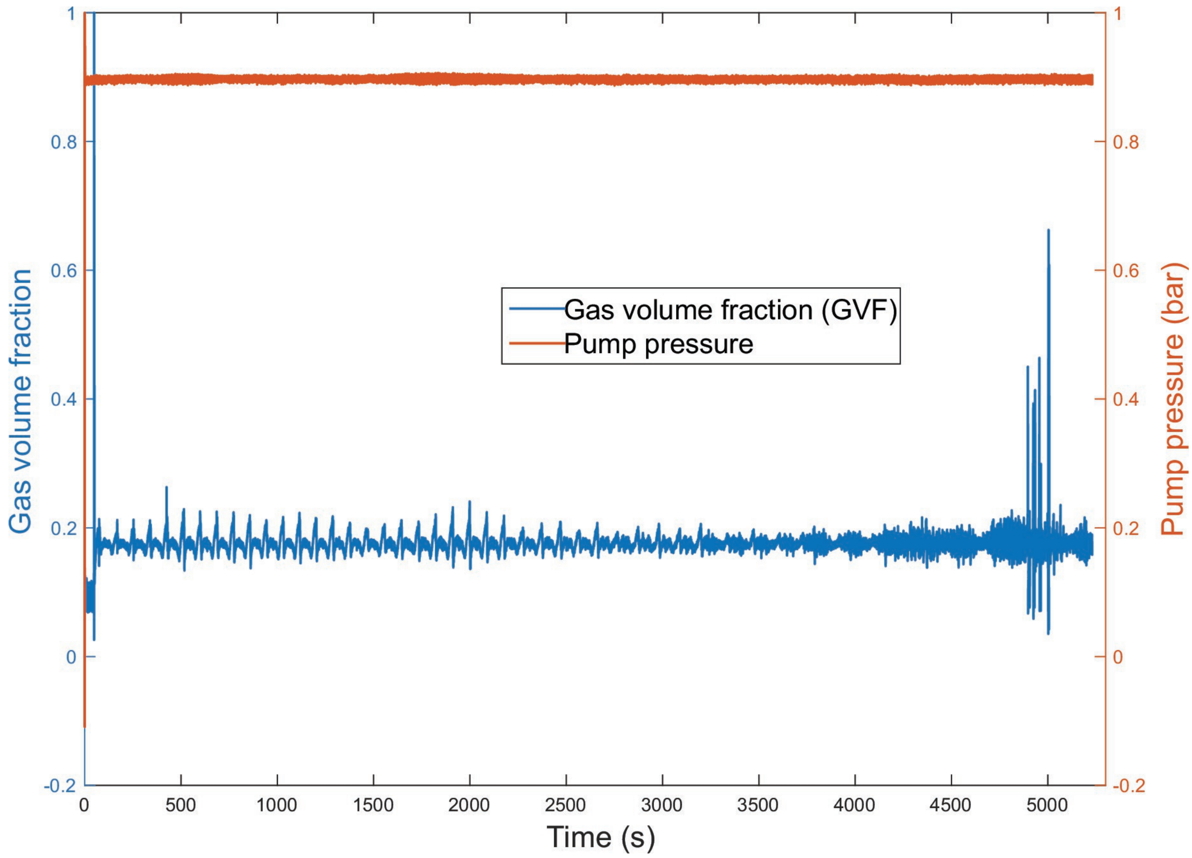

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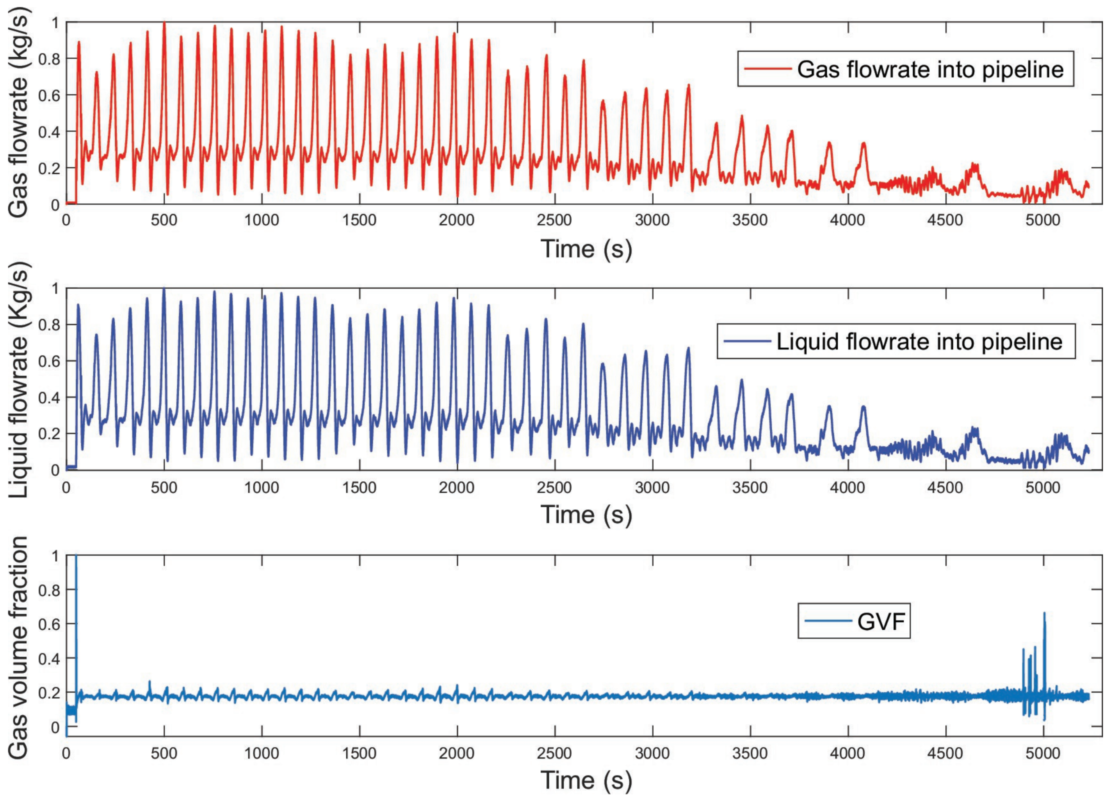

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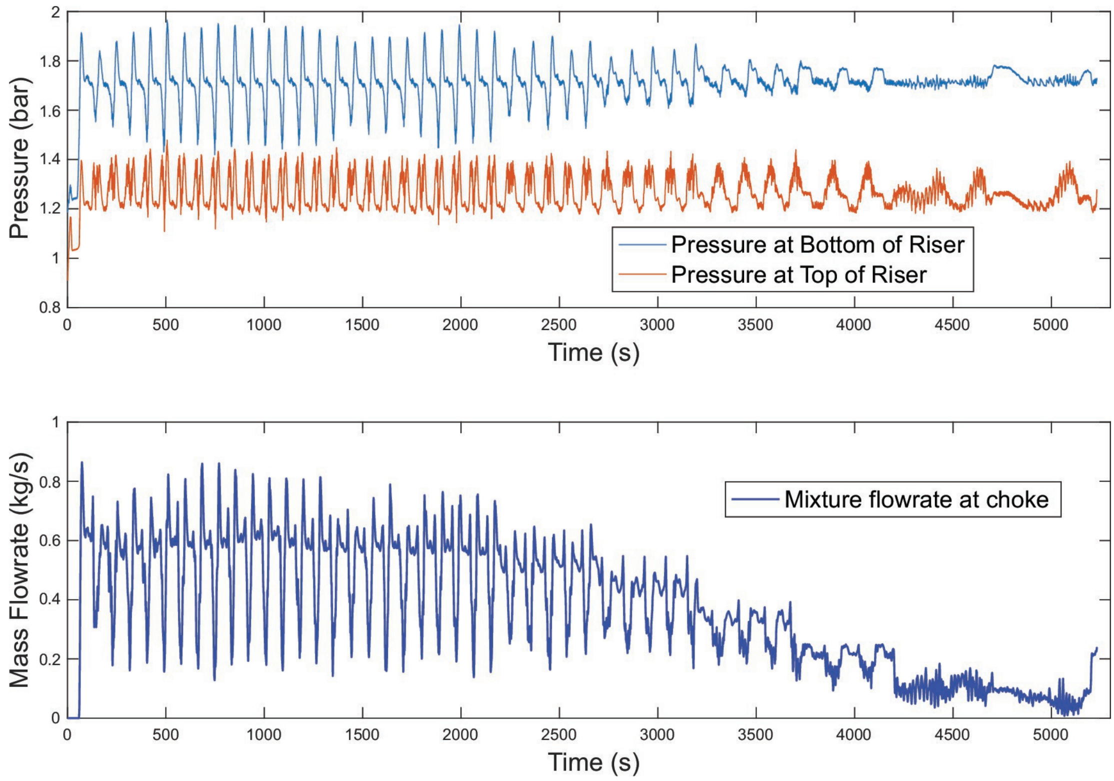

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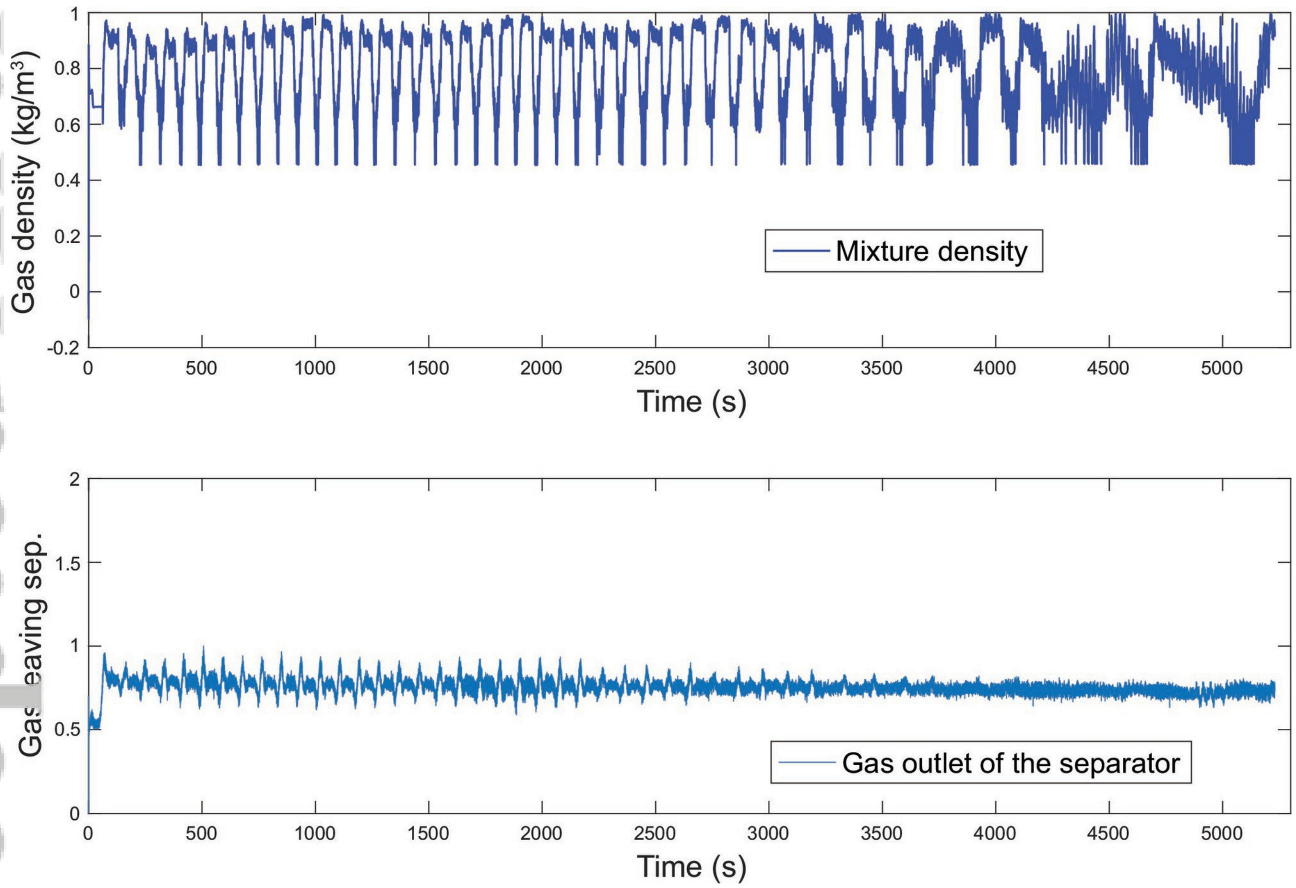

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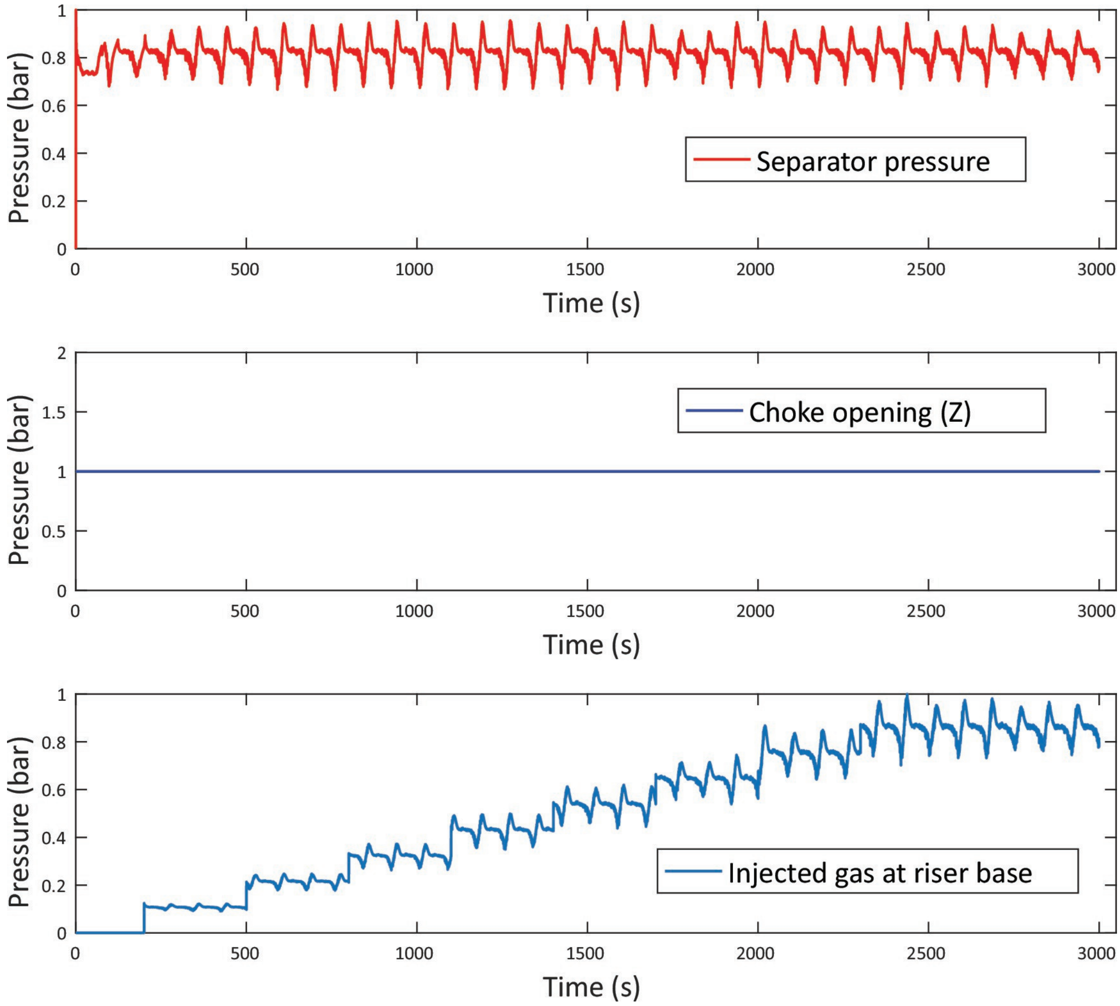

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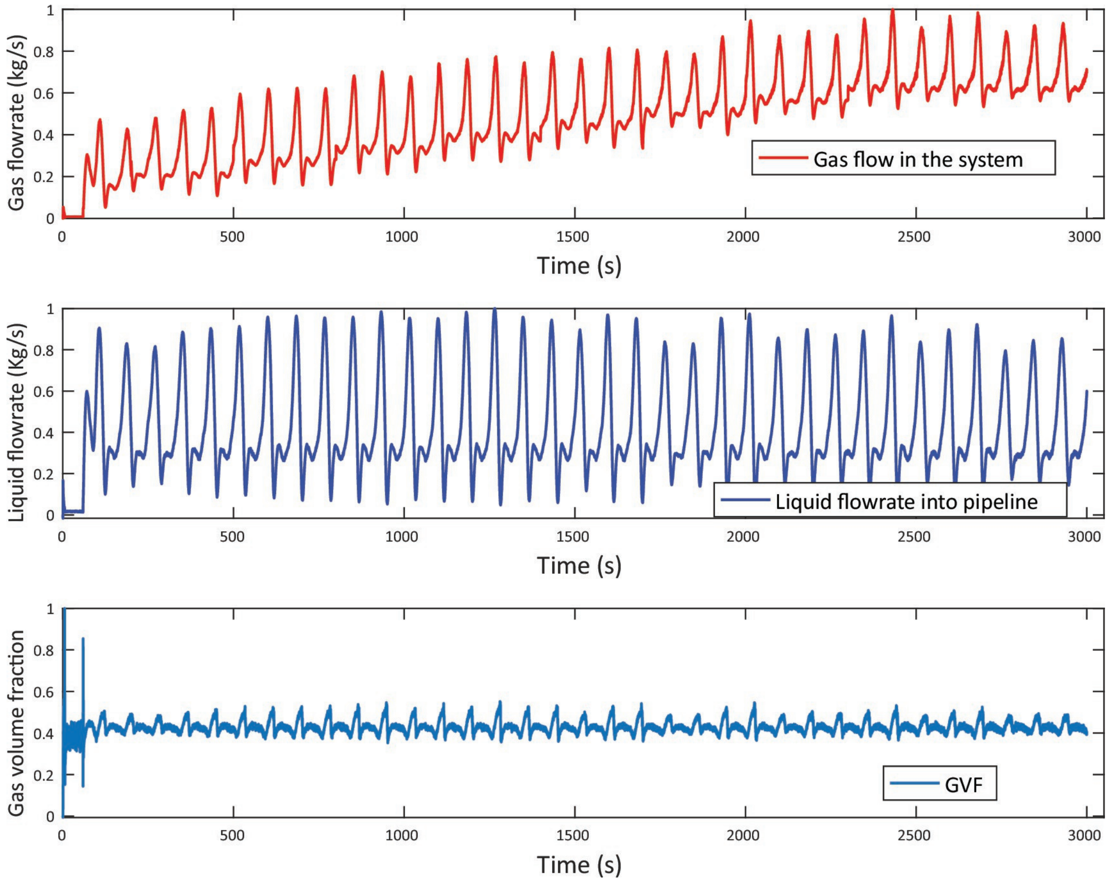

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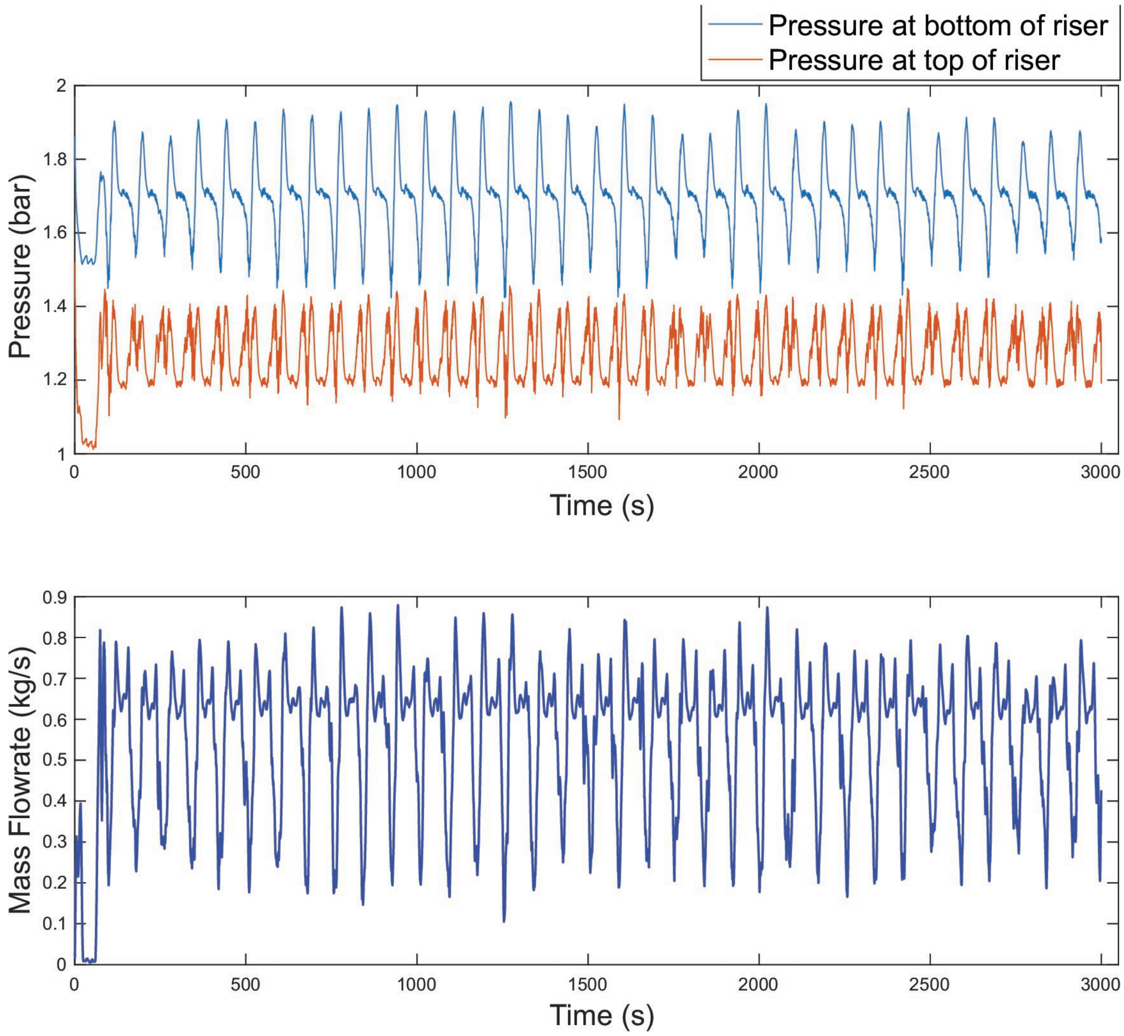

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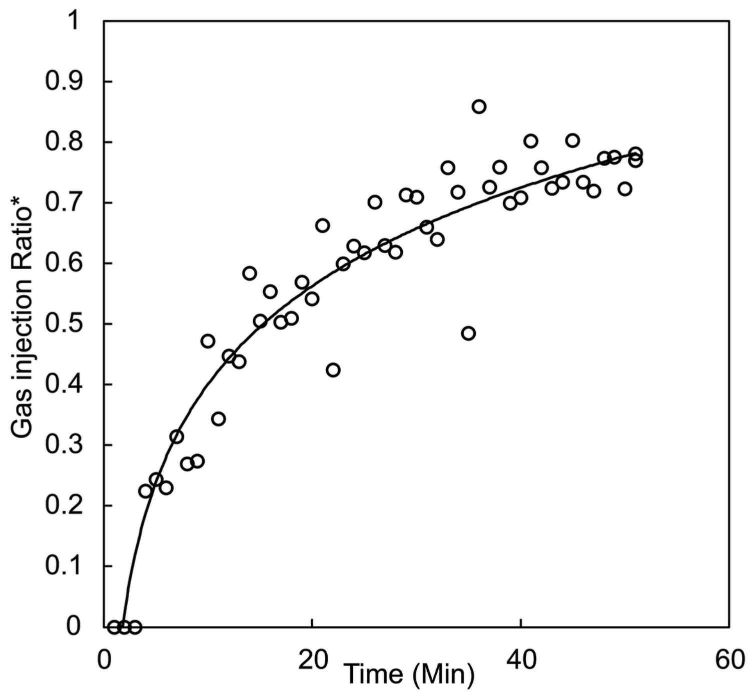

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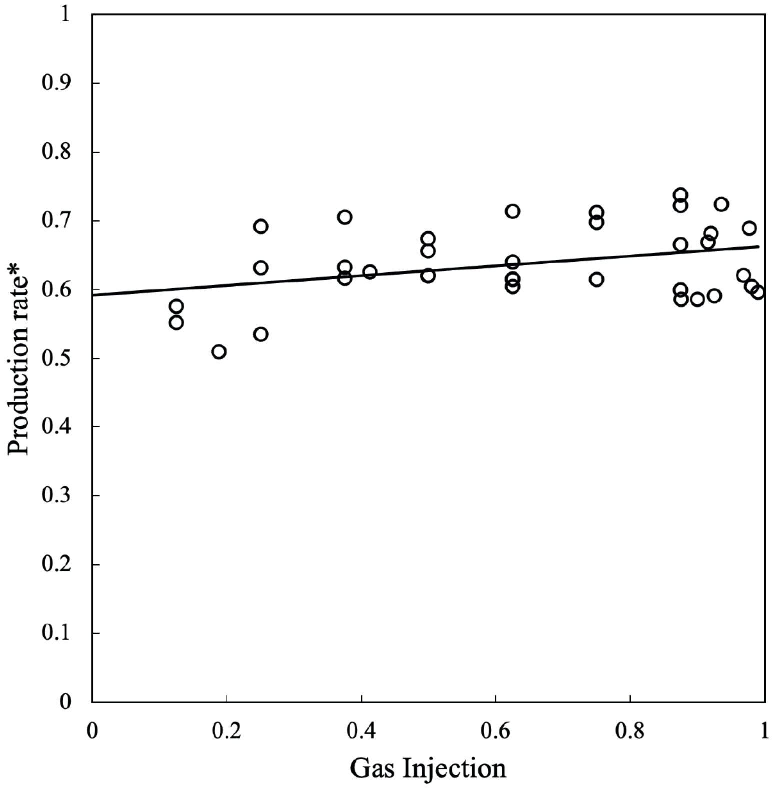

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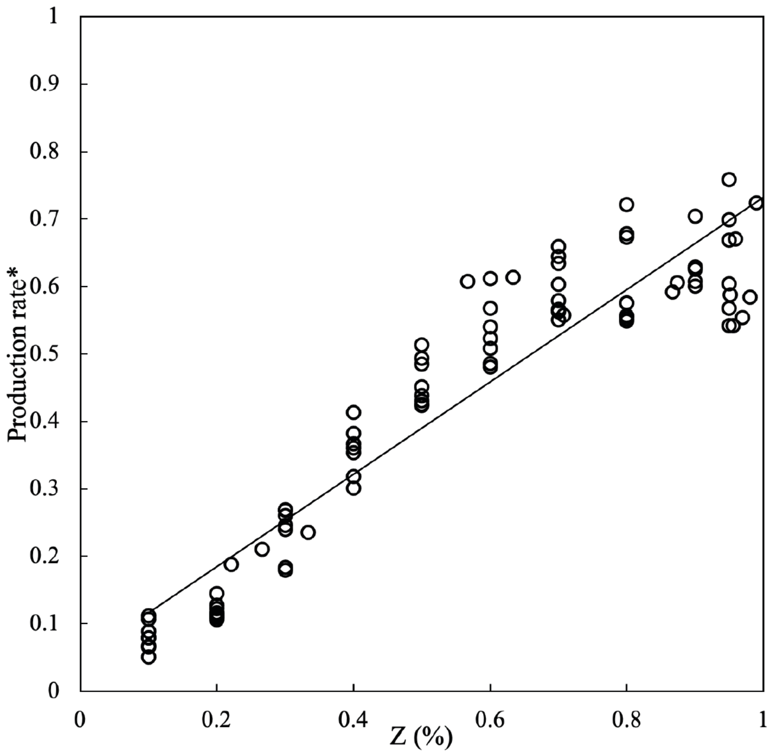

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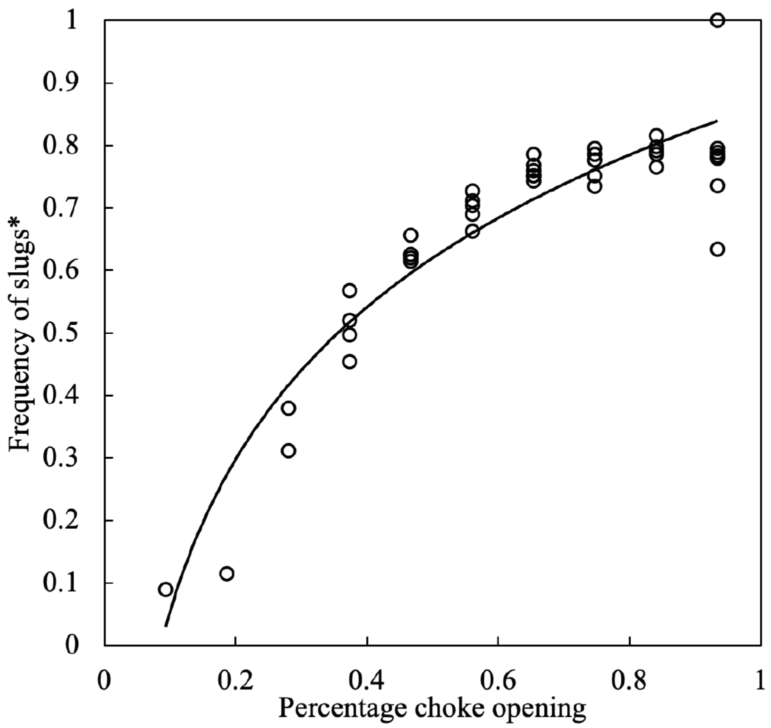

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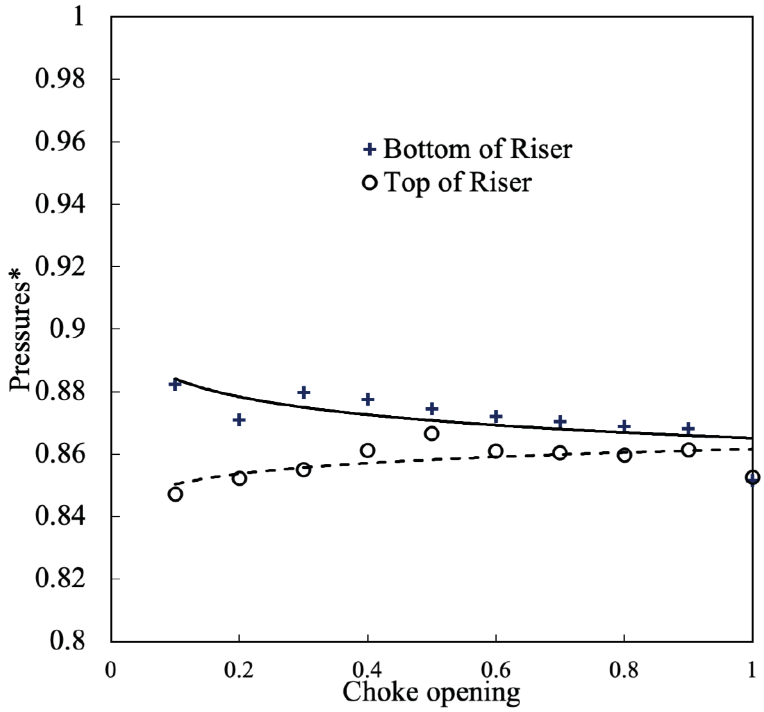

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